

BECHTEL CORPORATION (A)

Process Analysis and Optimization of a Desalting Plant Design

Mr. Rufus Crawford, a young chemical engineer, had been hired by the Bechtel Corporation in 1964 to assist in process analysis and cost studies of large multi-stage flash saline water conversion plants. An earlier Bechtel study was sponsored by the Office of Saline Water (OSW 116) to determine how much costs for an M.S.F. plant could be reduced by increasing the size of the operation and using exhaust steam from a gas-fired power-generating station to heat the sea water.

Rufus said that they had tried various optimization techniques using programs they had obtained from the National Computer Library. He said, "It's hard to pick up the logic of those library computer programs without spending a great deal of analysis time."

Thus, Crawford was enthusiastic when Dr. Lionel S. Galstaun and Mr. Richard L. Clark called him into Dr. Galstaun's office to ask him if he would like to work with Dr. Douglas Wilde of Stanford in the development of a new optimization program.

The Water Resources Section of this large engineering and construction firm in which Crawford worked had just received management approval for an in-house sponsored project to study the feasibility of large scale flash distillation in order to (1) improve the company's competence in desalting technology and (2) develop experience in computer-aided optimization.

Although the company was well experienced with desalting on a pilot plant scale, large plants would be something new. Tradeoffs to be made would include such things as energy required versus investment in heat transfer area. What temperatures and pressures to operate at, and what kinds of water treatment (sulphuric acid or phosphate additives?) to use for suppression of scaling were among the questions in need of answers.

Crawford saw his opportunity to really answer some of those questions he had been asking himself about those computer programs and gladly accepted the assignment.

Prepared at Stanford University during the 1966 National Science Foundation Summer Institute conducted by the Design Division of the Mechanical Engineering Department. The initial contact with Bechtel was made by Karl H. Vesper. The case was prepared by Joseph W. McCutchan, U.C.L.A., Robert P. Vail, University of Arizona, and Edward N. Ziegler, Polytechnic Institute of Brooklyn, participants in the Summer Institute. The assistance of Mr. Rufus Crawford was indispensable since the case really relates his experience. The cooperation of Mr. James Appleton and the permission of the Bechtel Corporation to use this case for educational purposes is greatly appreciated.

THE BECHTEL CORPORATION (A)
Applied Technology Group of the
Scientific Development Department

Bechtel is recognized in the United States as a builder for industry and government, and for its many years as an international engineer, contractor and project manager of facilities for the natural resources and energy industries. Under single responsibility the Company provides complete service for the realization of projects, including economic and technical studies, engineering design, procurement, and construction or project management. During early years the organization participated in construction milestones as Hoover Dam and the San Francisco-Oakland Bay Bridge. More recently, activities have expanded to encompass such fields as nuclear engineering, refinery systems, food processing, chemical plants, pipelines, ore processing and other basic private industries.

In the public sector the firm is active in water use and conservation, rapid transit, water desalination and the missile and space programs.

Indicative of its concern with technological advancement, Bechtel has been a pioneer in the application of nuclear energy. Starting with construction of EBR-1 in 1949, the first nuclear reactor in the world from which useful electric power was produced, the company has been a major participant in the conception, evaluation, design, engineering and construction of many nuclear projects, including fourteen generating plants.

In recent years when the idea of combining nuclear power with a sea water desalination plant was proposed, Bechtel was one of the engineering firms that took an active interest. Bechtel had long been acquainted with the use of processes for desalting makeup sea water for power plants. The concept of large-scale nuclear power and desalting combines was a technically attractive prospect. Bechtel had designed and built the Mandalay Steam Power Station near Oxnard, California. This was the first power plant in the U.S. to utilize the multistage flash distillation process to produce the total fresh water requirement for the plant. The plant had 100,000 gpd capacity and was designed and built on a subcontract by Cleaver Brooks, Inc.

Background on the Choice of the M.S.F. Process

When Dr. Lionel S. Galstaun came to work for Bechtel's Scientific Development Department in 1962, he further stimulated interest in the combined power-water concept and had a keen personal interest in the multistage flash process. Soon after that date Bechtel entered into a contract with the Office of Saline Water to make a comparative study of

the three most successful government pilot plants. The study had to be based on a normalized cost estimate so that the costs of the individual facilities would be comparative. Considerable criticism existed at that time because it was felt by many engineers that cost estimates being made by many different engineering concerns were not truly comparable.

That study, published in 1963, was entitled "A Study of Large Scale Saline Water Conversion Plants" and contained the following cost of product water summary (in cents per 1000 gal.)*

Capacity Million GPD	0.25	1.0	2.0	8.0	14.0
Electrodialysis**	140.0	115.0	95.9	---	---
Multiple effect (LTV)	---	146.3	118.7	85.5	---
Multistage flash (MSF)	---	130.2	105.2	73.0	64.8

This study based on the three demonstration plants then in operation further confirmed the belief that the MSF process held an increasing cost advantage as the size was increased. In fact, it was suggested that a 350 day/yr plant availability would be justified for the multistage flash process instead of the normalized 330 day used in the study.

A picture of the 36 stage San Diego Plant is shown in Exhibit 1. Exhibit 2 is a flow diagram for 50,000,000 gpd based on an extrapolation of this idea. Exhibit 3 shows a typical cutaway evaporator vessel for the 50,000,000 gpd design concept.

* "A Study of Large Size Saline Water Conversion Plants" by the Bechtel Corporation made for the Office of Saline Water, U.S. Dept. of the Interior. R & D Report 72, March 1963.

** The electrodialysis plant was based on brackish water and thus eliminates it from our consideration when operating on sea water and above 2 million GPD capacity.

Events Leading to Crawford's Assignment

In January 1965 a member of the Scientific Development Department of Bechtel attended a course in Dynamic Programming at Stanford University taught by Dr. Douglas Wilde. Upon reporting this experience to Dr. Galstaun, Dr. Galstaun suggested that it would be a good idea for Bechtel to retain Dr. Wilde, on a consulting basis, to develop generalized optimization approaches to process design problems. Dr. Galstaun suggested that process optimization would be of interest to the Process Engineering and Water Resources groups of the Applied Technology Section. Dr. Galstaun further suggested that an initial study be made to see whether new optimization approaches could be applied to the MSF desalination process.

Thus, the initial design team consisted of Mr. R.L. Clark, Mr. Rufus Crawford, and Dr. Wilde with Dr. Galstaun supervising the study. Mr. Clark was a senior member of the desalination group and served as the direct supervisor for Rufus Crawford. Many of the decisions needed for Crawford to get started were made by Clark and Galstaun based on prior desalination feasibility studies carried out by the Bechtel Corporation.

It was in June 1965 that Dr. Wilde was actually retained as a consultant and a start was made to optimize the MSF process utilizing a computer program. Soon after this date, Dr. Galstaun was assigned to the New York office of Bechtel and Mr. R.L. Clark was assigned to the office in Mexico City. The working team then became:

Mr. James Appleton, Project Supervisor
Mr. Rufus Crawford, Study Engineer
Dr. Douglas Wilde, Consultant Chemical Engineer
Mr. Moredicai Avriel, Stanford Graduate Student, Summer employee
of Bechtel and assistant to Dr. Wilde.
Miss Marilyn Martin, Computer Programmer
Mr. John Chapman, Computer Programmer/Chemical Engineer

Crawford recalled the comparative study made by Bechtel on the three OSW demonstration plants. Crawford said that for convenience the MSF process could be called the long tube horizontal (LTH) process in contrast to the long tube vertical (LTV) process. He said, "The processes do have a number of differences other than the fact that the tubes are horizontal or vertical." The LTV was the only other distillation plant considered in the comparative study. The one million gallon per day plant at Freeport, Texas, is of the LTV type and operating data are available. In this multiple effect distillation process sea water enters the first effect and is heated by an external source. At Freeport this is a steam supply from the immediately adjacent Dow Chemical plant. Steam that is formed in the first effect is condensed in the second effect which evaporates more sea water (at a reduced pressure). This process is repeated in succeeding effects, the pressure changes corresponding to the temperature difference required to produce the flow of heat needed. See Figure 1.

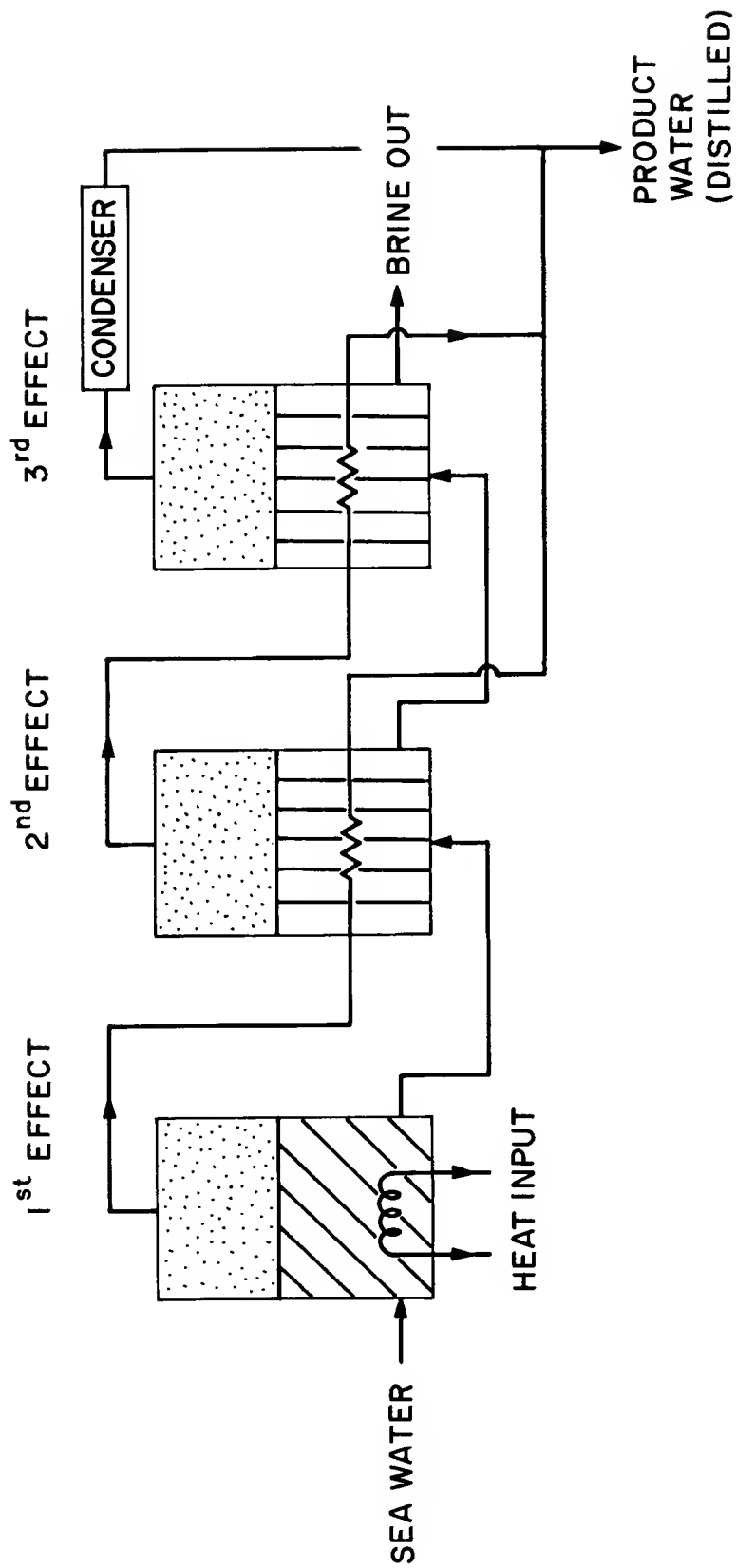


Figure 1

In optimizing the MSF process, Crawford said their first inclination was to apply the principles of dynamic programming, a promising optimization technique, directly to the MSF process but they found that the computer program required too many storage cells. They also felt a need to find a more efficient optimization technique capable of analyzing the many MSF system parameters. They knew they were going to take partial derivatives and this ruled out the trial and error solution analysis even though a cycle analysis could have been performed on that basis.

Dr. Wilde, after studying the MSF process for about one week suggested that the Pontryagin Maximum Principle (derived in Russia) might be applicable to this problem and prove to be a useful tool in optimizing this system. This technique is a sophisticated application of variational calculus and had been investigated by Dr. Wilde at Stanford. However, obtaining the process variables in the proper form for the computer program proved to be a challenging assignment for Crawford and Avriel.

"Gradually we realized," Rufus said, "that our procedure could be divided into four stages." Rufus, himself, would develop the mathematical MSF model suitable for the computer. Wilde and Avriel would be responsible for setting up the algorithm, i.e., the mechanics of the maximum principle concept. The team as a whole planned to work on the adjustment procedures for improving the MSF model and finally, Rufus would evaluate the output of the computer program and the effectiveness of alternatives. The cost of water per 1000 gallons was to be the basic parameter to be optimized.

The problem which was defined by Crawford in consultation with Galstaun and Clark was to have the following initial boundary conditions:

1. The MSF system would have greater than 10 million gpd capacity.
2. Sulfuric acid or polyphosphate additives were to be used for scale control but not to be a major optimization parameter.
3. A single effect recycle system would be used consisting of:
 - (a) heat input section
 - (b) heat recovery section
 - (c) heat rejection section

These general boundary conditions were selected so as not to over-restrict the optimization study. Mr. Crawford had available the following simplified flow diagram for a flash plant using polyphosphate additive.

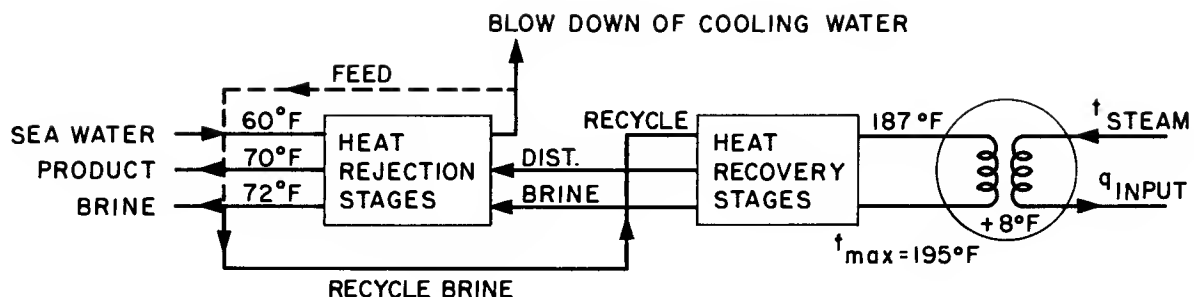


Figure 2. Simplified Diagram of MSF Process

This cycle operated reliably at San Diego and other site locations throughout the world. It produced one million gallons per day when using the polyphosphate (HAGEVAP LP at 4 ppm) additive to prevent the adherence of scale (mineral deposits) to the heat transfer surfaces. Then with a minimum of change they were able to produce 1,400,000 gallons per day using a maximum brine temperature of 250°F. In this case, scale was controlled by the addition of (125 ppm) a stoichiometric amount of H_2SO_4 to break down the HCO_3 ion present in sea water. The evolution of CO_2 after acidifying with H_2SO_4 requires that the deaeration capacity of the system be double that for a system which removes the dissolved gases alone.

The range of heat input "q" for multistage flash systems varies from 50 to 200 Btu per pound of product water, with values near 100 being more common in current practice. Another useful parameter, the gained output ratio (g.o.r.) is defined as $\frac{1000}{q}$ where the 1000 is approximately the latent heat of vaporization. For a typical one million gpd plant such as San Diego

$$(g.o.r.) = 10.5 = \frac{1000}{95.2} \frac{(\text{Btu/lbm steam})}{(\text{Btu/lbm distillate})}$$

Some other conditions had to be specified in order to analyze the "phosphate" cycle specifically. The steam supplied to the heater was assumed to be supplied at 213°F. The brine entering the heater was 187°F and the other assumed conditions are shown on the simplified flow diagram (Figure 2).

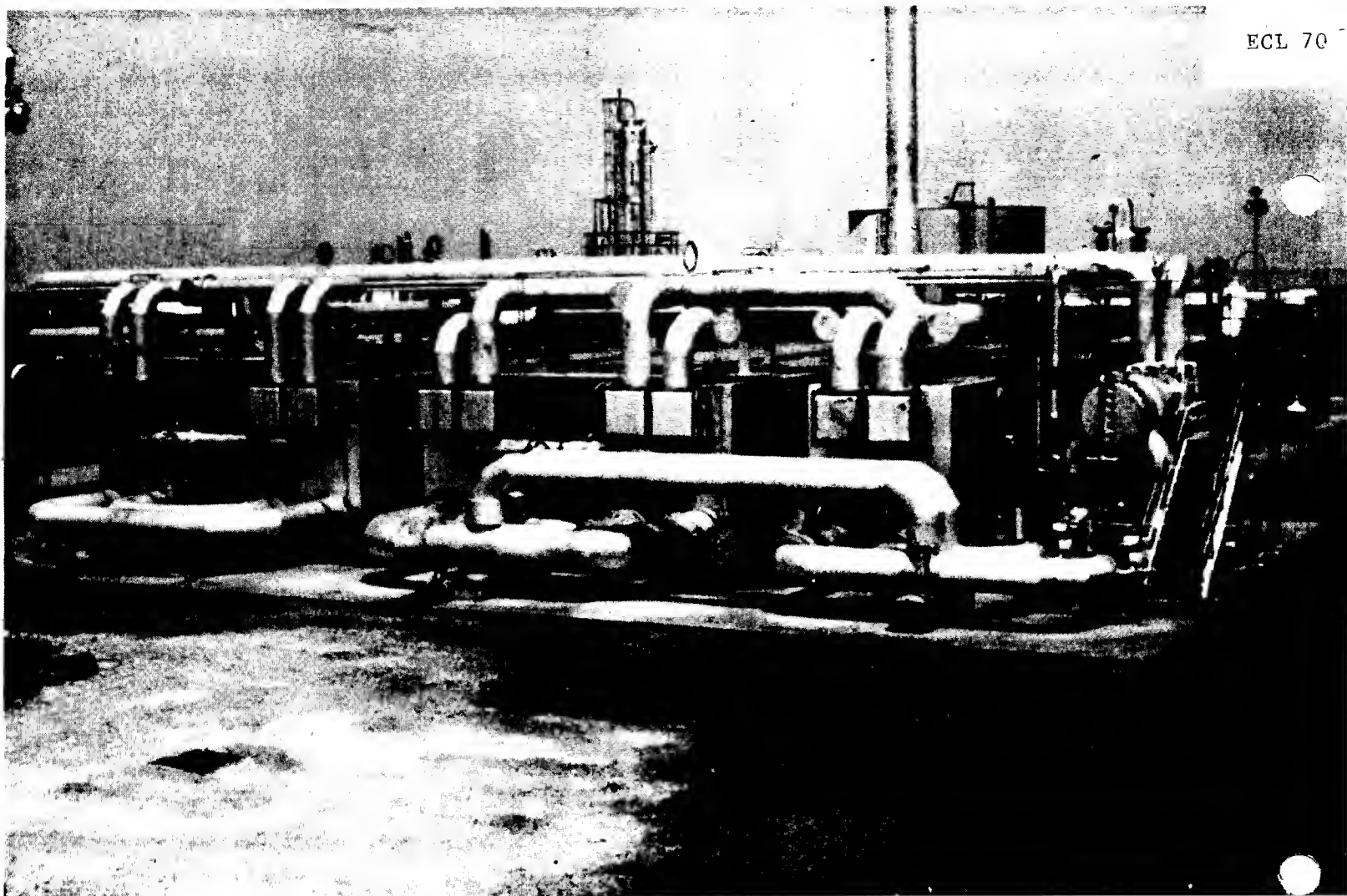
The concentration factor is defined as the ratio,

$$C.F. = \frac{(\text{ppm}) \text{ salts in the blowdown}}{(\text{ppm}) \text{ salts in the sea water}}$$

for MSF cycle values are generally between 1.7 and 2 as to prevent scale build-up on the heat transfer surfaces. The number of stages may be specified. A typical value for a g.o.r. of 6 might be 25 stages while for a g.o.r. of 12 approximately 50 stages would be used.

The fluid velocities found within the condenser tubes are limited by inlet and outlet erosion problems. Depending on the tube material the fluid velocity in the tubes may be taken as 4 ft/sec to 7.5 ft/sec. These velocities have to be set for all three sections. The velocities are needed to estimate the heat transfer coefficients and ultimately the condenser area needed for each of the plant sections.

Crawford knew that he must derive the process analysis equations before Avriel and Dr. Wilde could start their optimization program. Some method would have to be developed to measure the effectiveness of this program. He was mentally planning his approach as he drove home.

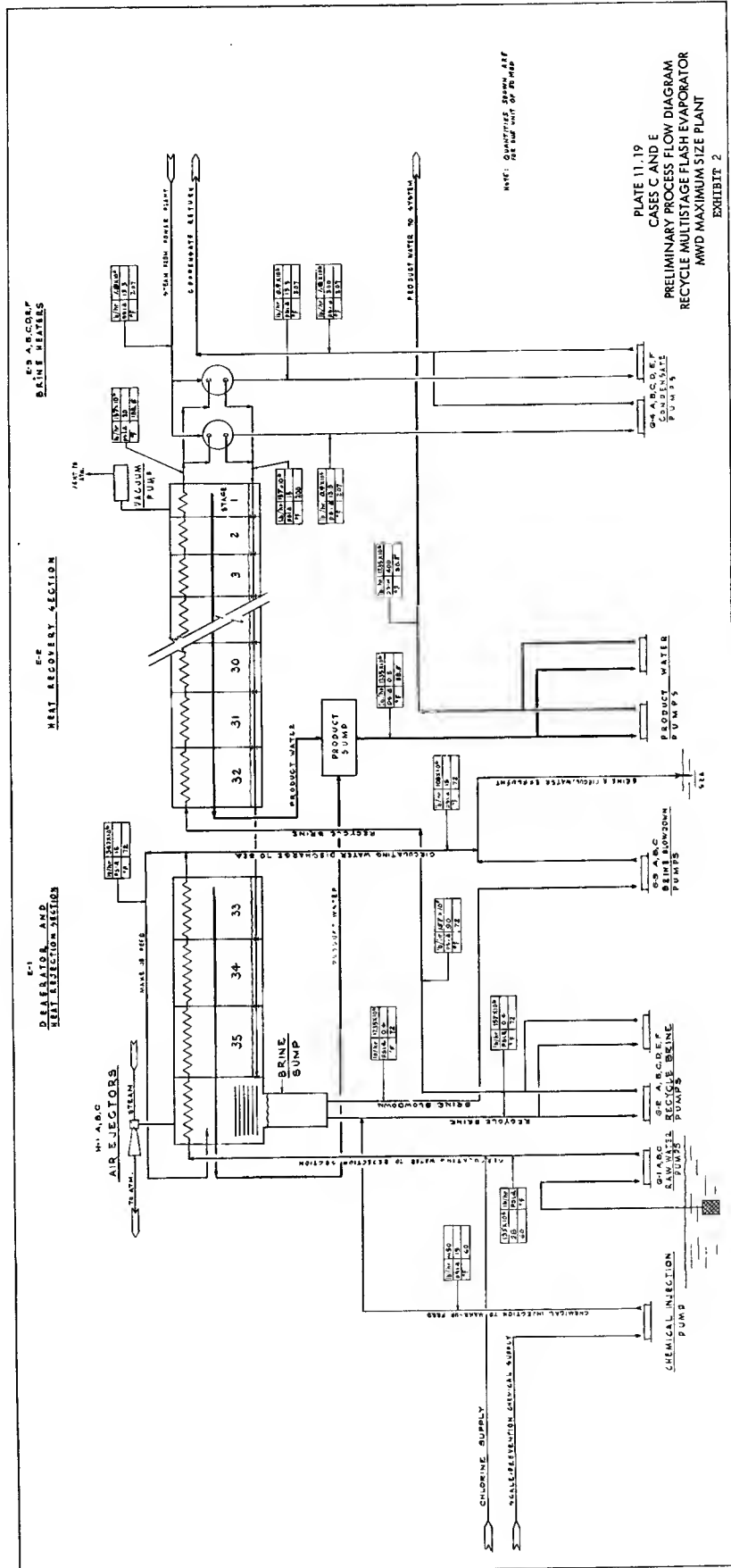


36-STAGE LTH DISTILLATION PLANT

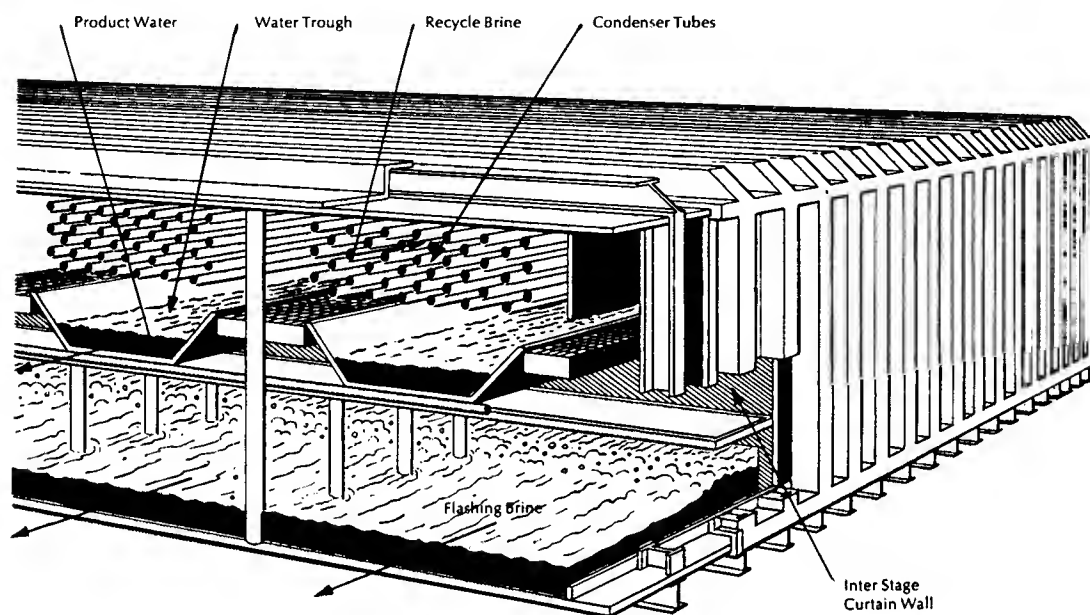
Office of Saline Water
San Diego, California

This 36-stage, 1,000,000 gpd flash facility began operation as a demonstration plant in late 1961. The Bechtel Corporation was selected in 1962 by the Office of Saline Water to evaluate the plant's performance, and to make recommendations, and determine the cost of water. Seen above are the ends of the five flash vessels, the brine heater, and, in the background, the boiler feed deaerator and boiler stack. Since this picture was taken, this plant was dismantled and shipped to Guantanamo Bay, where it is now producing water to meet the needs of the naval base.

EXHIBIT 1



As shown in the following cutaway view of a typical evaporator vessel, each vessel contains condensing tubes, product water troughs, and a flashing brine area. The condensing tubes, carrying the recycle brine, extend through the internal walls, thus going from stage to stage in each vessel. The flashing brine flows in the opposite direction from stage to stage through orifices in the vessel internal walls. The product water troughs for each stage are connected by loop seals and orifices.



Typical Evaporator Vessel

EXHIBIT 3

THE BECHTEL CORPORATION (B)

Process Analysis and Optimization of a Desalting Plant Design

Mr. Crawford decided it would be necessary to set up a mathematical model for a single stage as the first phase of an overall flow stream model. After the basic energy balances were determined for the single stage Rufus and Mordecai would extend their analysis to combined stages and finally to the MSF system as a whole. They expected to select variables suitable for the optimization scheme.

Single Stage Model

In the Simplified Flow Model of the i^{th} Stage, Avriel and Crawford

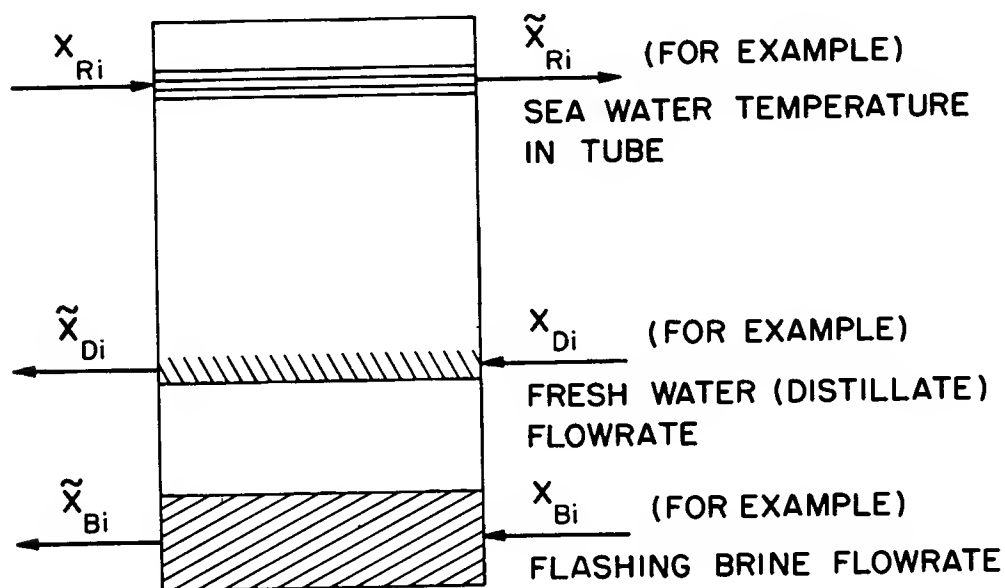


Figure 3 Simplified Flow Model of the i^{th} Stage

let X denote any property of a stream entering the stage and \tilde{X} that same property as the stream leaves. The subscripts R, D and B represent recycle brine, distillate and flashing brine, respectively. Crawford said that Dick Clark didn't like the \tilde{X} (\tilde{X}) symbol, saying "What happens if you lose a "tilde"? Actually a single symbol could have been used with the subscript i indicating the interface between stages instead of the stage number itself. However, the program was already underway so the notation was not changed.

THE BECHTEL CORPORATION (B)

Process Analysis and Optimization of a Desalting Plant Design

Mr. Crawford decided it would be necessary to set up a mathematical model for a single stage as the first phase of an overall flow stream model. After the basic energy balances were determined for the single stage Rufus and Mordecai would extend their analysis to combined stages and finally to the MSF system as a whole. They expected to select variables suitable for the optimization scheme.

Single Stage Model

In the Simplified Flow Model of the i^{th} Stage, Avriel and Crawford

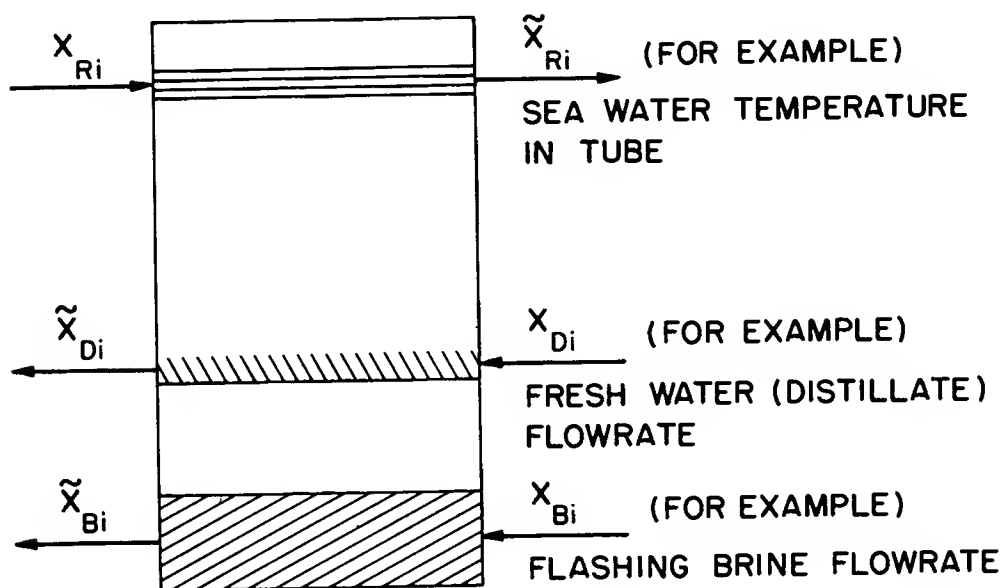


Figure 3 Simplified Flow Model of the i^{th} Stage

let X denote any property of a stream entering the stage and \tilde{X} that same property as the stream leaves. The subscripts R, D and B represent recycle brine, distillate and flashing brine, respectively. Crawford said that Dick Clark didn't like the \tilde{X} (\tilde{X}) symbol, saying "What happens if you lose a "tilda"? Actually a single symbol could have been used with the subscript i indicating the interface between stages instead of the stage number itself. However, the program was already underway so the notation was not changed.

Since each stage is assumed to be at a constant pressure, (flashing takes place at the stage interface as the brine is throttled through an orifice) the relationships between stream temperatures may be deduced from an enthalpy balance. Thus the enthalpy lost over the stage by the brine stream must equal the enthalpy gained by the distillate and recycle brine streams. Crawford made the decision as to which brine stream heat capacity, density, thermal conductivity, viscosity and boiling point elevation data would be used in the calculations. He mentioned that it was accepted practice by now to use the data from the "Saline Water Conversion Engineering Data Book" compiled by the M. W. Kellogg Company under contract with the Office of Saline Water of the U. S. Department of the Interior. Crawford did not check these data directly but the known thermodynamic properties of fresh water and extensive data on NaCl solutions serve as a rough check. These data may be found in Exhibit 4 originating from Oak Ridge National Laboratory. Crawford pointed out that, "The method of solution is independent of the data fed and a higher degree of accuracy can be achieved as more reliable data become available." For purposes of checking the single stage heat balance equations the variations in brine properties with concentrations was unnecessary.

Combined Stage Model

After completing and checking the single stage analysis, Crawford and Avriel expanded their concept to include the entire heat recovery section (see Exhibit 5) and with Miss Martin's assistance programmed the equations. They analyzed a 33 stage operation so that a check might be made with an earlier report, Bechtel No. 4427 which was issued by the Office of Saline Water as R. & D. Report No. 116 and may be purchased from the Superintendent of Documents, U. S. Department of Commerce. This report used a different computational sequence and was somewhat lengthier. This comparison saved them the job of making a hand calculation to check the computer solution. The results of the analysis showed that the new relations used by the computer appeared to be consistent with those used in 4427 but a final judgment was not made at this time since the heat transfer coefficients were slightly different.

The next step was to write a heat balance on a single stage of the heat rejection section. The rejection stages were then combined. The brine heater was then included assuming inputs from a steam source of $+8^{\circ}\text{F}$ above the maximum allowable brine temperature. Crawford said that the program will accept any ΔT but that Mordecai did a detailed analysis of this using geometric programming and concluded that 8°F was very near the optimum for this specific application. The recovery, rejection and brine heater sections were then combined to form an integrated system to which boundary conditions could be applied.

The Avriel letter (Exhibit 5) records the heat and material balance relations of the MSF study. Crawford said, "I would have liked this material to have been better organized for the record, but I agree completely with the conclusions." In fact, he felt he was a part of those conclusions.

Part of the study had been to derive equations for the heat transfer coefficients. Avriel used the standard textbook equations such as the Dittus-Boelter equation. These had been used by Bechtel prior to the study and had been shown to reproduce with reasonable accuracy heat transfer data taken from the San Diego and other facilities. They were eager to get the results of the first computer run since they were not 100% sure that they hadn't made some mistake in their derivations and wanted to check out the program. They were using report No. 4427 but they could have used actual operating data such as that of the San Diego Plant (Exhibit 6).

The cycle analysis was recognized as an application of the first law of thermodynamics and since this is true by definition, what goes in should balance what comes out. However, there was an error somewhere because the computer program repeatedly gave an imbalance. This showed up as a predicted sea temperature of 61° F instead of the 60° F temperature originally postulated. The cycle balance was accomplished by writing equations for the first stage (stage nearest the heater); then the outputs from this stage were fed to the second stage until a balance was achieved then fed to the third, etc. The final values obtained in this manner should have checked with an overall balance for the system. However, they did not and the error indicated that more heat was going in than was coming out. In order to place a high degree of confidence in the program the inconsistency had to be resolved. Mr. Crawford saw three possible sources of error. First, his hand calculation might have been in error. Or, the basic equations may not have been set up correctly. Finally, some error might have occurred in the computer program.

Mr. Crawford started the tedious task of running down this inconsistency. He rechecked his calculations using a slide rule but soon realized that the significant figures given by the slide rule would yield a greater variation than the discrepancy for which he was looking.

Next he checked his equations and could find no error. He questioned the accuracy inherent in the method of calculation and concluded that this could account for only 1/4° F in the sea temperature which was the final output number. He then started to check the computations stage by stage on an electric calculator, using eleven numbers on the readout register so that the question of significant figures would not arise again.

In order to make the length of this task reasonable and because there was a question in Crawford's mind as to the influence of the empirical equation for the specific heats, Crawford assumed all specific heats to be

1.0 $\frac{\text{Btu}}{\text{lb}}$ and reran the computer program with this constraint. By the time

Crawford got through the heat recovery section the numbers were still checking the computer program. However, in going through the heat rejection stages the error reappeared so he concluded the error must be in that part of the program.

Avriel and Wilde were also searching for the error and they had about concluded that the problem was overspecified and had gone to Dr. Galstaun to discuss this possibility with him. Almost at this exact time John Chapman called to say the program had just run.

It appeared that the error had been in the equation for computing the constant A_{RN} . It seems the equation had been written in the program as

$$A_{RN} = \alpha_1 W_R \bar{C}_p \beta_i \quad \text{instead of} \quad A_{RN} = \alpha_1 W_{RN} \bar{C}_p \beta_i$$

where W_{RN} is the circulating cooling sea water in the heat rejection section and the computer program was calling for W_R , the circulating water in the heat recovery section.

Student Assignment

- (1) Set up the equations for heat and mass balance for the " i^{th} " stage in the heat recovery section. (as Crawford did)
- (2) Set up the equations for heat and mass balance for the " i^{th} " stage in the heat rejection section. (as Crawford did)
- (3) How should Crawford now proceed?

Exhibit 4Physical Properties

Physical properties will be calculated from the following relations (Oak Ridge)

1. Boiling Pt. Elevation (Low Concentration)

Concentration range: .02 - .12 wt. fraction, solids

Temperature range: 77 - 392°F

$$\Delta(^{\circ}\text{F}) = A1 + A2(C) + A3(T) + A4(C)(T) + A5(C)^2 + A6(T)^2 + A7(C)(T)^2 + A8(C)^2(T) \\ + A9(C)^3 + A10(T)^3 + A11(C)(T)^3 + A12(C)^3(T) + A13(C)^2(T)^2 + A14(C)^2(T)^3$$

where

C = wt. fraction, solids

T = temp., °F

$$A1 = 0.1014380 \text{ E-1}$$

$$A2 = 0.1021815 \text{ E+2}$$

$$A3 = -.6386588 \text{ E-4}$$

$$A4 = 0.6118005 \text{ E-1}$$

$$A5 = 0.8809554 \text{ E+1}$$

$$A6 = 0.2386759 \text{ E-6}$$

$$A7 = 0.2214495 \text{ E-4}$$

$$A8 = 0.1714722$$

$$A9 = 0.7173776 \text{ E+2}$$

$$A10 = 0.1170617 \text{ E-9}$$

$$A11 = 0.1042795 \text{ E-6}$$

$$A12 = 0.3620461$$

$$A13 = 0.7823281 \text{ E-4}$$

$$A14 = -.5218751 \text{ E-6}$$

2. Boiling Pt. Elevation (High Concentration)

Concentration range: .12 - .28 wt. fraction, solids

$$\Delta(^{\circ}\text{F}) = A1 + A2(C) + A3(T) + A4(C)(T) + A5(C)^2 + A6(T)^2 + A7(C)(T)^2 + A8(C)^2(T) \\ + A9(C)^3 + A10(T)^3 + A11(C)(T)^3 + A12(C)^3(T) + A13(C)^2(T)^2 + A14(C)^2(T)^3 \\ + A15(C)^3(T)^2$$

where

C = wt. fraction, solids

T = temp., °F

$$A1 = -.1441403 \text{ E+1}$$

$$A2 = 0.3791963 \text{ E+2}$$

$$A3 = 0.1081565 \text{ E-1}$$

$$A4 = -.1253869$$

$$A5 = -.1679212 \text{ E+3}$$

$$A6 = -.4486056 \text{ E-4}$$

$$A7 = 0.7590168 \text{ E-3}$$

$$A8 = 0.1200749 \text{ E+1}$$

$$A9 = 0.4579313 \text{ E+3}$$

$$A10 = 0.4010406 \text{ E-7}$$

$$A11 = -.4393590 \text{ E-6}$$

$$A12 = -.1505255 \text{ E+1}$$

$$A13 = -.3665256 \text{ E-2}$$

$$A14 = 0.1280709 \text{ E-5}$$

$$A15 = 0.5852867 \text{ E-2}$$

3. Specific Heat: $\text{Btu}/(\text{lb}_m)(^\circ\text{F})$

Data available for 40 to 250°F and 0 to 26% salt. The data was extrapolated to 400°F.

$$C_p = 1.0057733 - .13280442 \times 10^{-3} T - 1.5344907 C + .39090715 \times 10^{-2} TC \\ + .65092605 \times 10^{-6} T^2 + 2.4955446 C^2 - .11156771 \times 10^{-1} TC^2 \\ - .88941836 \times 10^{-5} T^2 C + .3598702 \times 10^{-4} T^2 C^2$$

where C - wt. fraction of salt

T - °F

4. Density: lb_m/ft^3

Data available from 40 to 212°F and 0 to 26% salt. The data has been extrapolated up to 400°F using data of pure water as a guide.

$$\rho = 62.717753 - .32152986 \times 10^{-2} T + .44315006 \times 10^{-2} C - .11647394 \times 10^{-1} CT \\ - .48932777 \times 10^{-4} T^2 + .16449945 \times 10^{-2} C^2 + .13761984 \times 10^{-1} C^2 T \\ + .34231326 \times 10^{-5} CT^2 - .92275301 \times 10^{-4} C^2 T^2$$

C - wt. fraction of salt

T - °F

5. Thermal Conductivity: $\text{Btu}/(\text{hr})-(\text{Ft})(^\circ\text{F})$

Data available from 40 to 250°F and 0 to 26% salt.

Data has been extrapolated to 400°F using purewater data as a guide.

$$k = -.0655C + .29556783 + .8794748 \times 10^{-3} T \\ - .23026302 \times 10^{-5} T^2 + .1609445 \times 10^{-8} T^3$$

c - salt conc.(wt.fraction), T-°F

6. Viscosity: $\text{LB}_m/(\text{hr})-(\text{Ft})$

Data available from 40 to 250°F and 0 to 26% salt. The data was extrapolated to 400°F

$$\mu = 22.216306 - 14.161033C - 16.460588 (\log_{10} T) + 16.430566 (\log_{10} T) C \\ + 284.84684 C^2 + 3.0949585 (\log_{10} T)^2 - 4.205263 C (\log_{10} T)^2 \\ - 219.40291 (\log_{10} T) C^2 + 42.287729 C^2 (\log_{10} T)^2$$

C - salt conc. (wt. fraction), T-°F

$$\mu = (\text{lb}_m/\text{hr}-\text{F}_T)$$



BECHTEL CORPORATION

INTER-OFFICE MEMORANDUM

Date July 19, 1965

To M. G. Martin

From M. Avriel

Dept. Scientific Development

Copies to L. S. Galstaun
D. J. Wilde
R. L. Clark
R. W. Crawford

At 301 Mission

Subject:

Programming Heat & Mass Balance Relations of
MSF Optimization Study

We are now ready to program and test the heat and mass balance relations for the optimization study. These relations determine the physical conditions at the outlet of each stage as a function of the physical conditions at the inlet to the same stage. Knowing the terminal conditions at one end of the flashing stages enables one to calculate recursively the conditions at each stage.

Let

H_{R_i} = enthalpy of recycle brine, entering stage i , (Btu)

H_{B_i} = enthalpy of flashing brine, entering stage i , (Btu)

H_{D_i} = enthalpy of distillate, entering stage i , (Btu)

\tilde{H}_{P_i} = enthalpy of distillate produced in stage i , (Btu)

T_{R_i} = temperature of recycle brine entering stage i , ($^{\circ}\text{F}$)

T_{B_i} = temperature of flashing brine entering stage i , ($^{\circ}\text{F}$)

EXHIBIT 5 - (Page 2 of 11)

- T_{Di} = temperature of distillate entering stage i, ($^{\circ}\text{F}$)
 W_{Bi} = flow rate of flashing brine entering stage i, (lb/hr)
 W_{Di} = flow rate of distillate entering stage i, (lb/hr)
 W_R = flow rate of recycle brine in heat recovery section
 C_p = specific heat (Btu/lb $^{\circ}\text{F}$)
 λ = latent heat of vaporization (Btu/lb)
 Δ = boiling point elevation ($^{\circ}\text{F}$)
 U = overall heat transfer coefficient (Btu/hr ft 2 $^{\circ}\text{F}$)
 n_i = number of tubes in stage i
 D_i = O.D. of tubes in stage i (ft)
 L_i = tube length in stage i (ft)

Also let

\tilde{X}_i denote a variable leaving stage i, or a physical property corresponding to the conditions of a stream leaving stage i. For example \tilde{C}_{PBi} is the specific heat of the flashing brine at temp and concentration as it leaves stage i.

\bar{X}_i denote an average variable throughout a stage unless specified otherwise.

The following basic relations exist at each stage of the recovery section:

Overall heat balance on stage i

$$(1) \quad H_{Di} + H_{Bi} + H_{Ri} = \tilde{H}_{Di} + \tilde{H}_{Bi} + \tilde{H}_{Ri}$$

Heat transfer at stage i.

$$(2) \quad \int_{T_{Ri}}^{\tilde{T}_{Ri}} W_R C_{PR} dT_R = \int_0^{L_i} U_i n_i \pi D_i (\tilde{T}_{Di} - T_R) dx$$

EXHIBIT 5 - (Page 3 of 11)

where it is assumed that the temperature of the distillate is constant across the stage and it is equal to the equilibrium temp.

Heat balance on flashing brine:

$$(3) \quad H_{Bi} = \bar{H}_{Bi} + \bar{H}_{Pi} + \lambda_i (W_{Bi} - \bar{W}_{Bi})$$

Define the following average quantities:

Average specific heat of distillate at stage i:

$$(4) \quad \bar{C}_{PDi} = \frac{C_{PDi} + \bar{C}_{PDi}}{2}$$

Average specific heat of recycle brine at stage i:

$$(5) \quad \bar{C}_{PRi} = \frac{C_{PRi} + \bar{C}_{PRi}}{2}$$

Average specific heat of flashing brine at stage i:

$$(6) \quad \bar{C}_{PBi} = \frac{C_{PBi} + \bar{C}_{PBi}}{2}$$

Note that (4) and (5) are functions of inlet and outlet temperatures, while (6) is a function of inlet and outlet temperatures and concentrations.

Weighted average overall heat transfer coefficient at stage i:

$$(7) \quad \bar{U}_i = \frac{U_i (\bar{T}_{Di} - T_{Ri}) + \bar{U}_i (\bar{T}_{Di} - \bar{T}_{Ri})}{(\bar{T}_{Di} - T_R) + (\bar{T}_{Di} - \bar{T}_{Ri})}$$

Average boiling point elevation at stage i:

$$(8) \quad \bar{\Delta}_i = \frac{\Delta_i + \bar{\Delta}_i}{2}$$

Eq. (8) is a function of both temperature and salt concentration

EXHIBIT 5 - (Page 4 of 11)

Let

$$(9) \quad \alpha_i = \left\{ \left(\frac{\tilde{\lambda}_i}{\bar{C}_{PBi} \bar{\Delta}_i} \right) - 1 \right\}$$

$$\beta_i = \left\{ 1 - \exp \left[\frac{\bar{U}_i n_i \pi D_i L_i}{W_R \bar{C}_{PRi}} \right] \right\}$$

Then, combining eqs. (1) through (9) one obtains:

$$(10) \quad \tilde{T}_{Di} = \frac{\alpha_i W_R \bar{C}_{PRi} \beta_i \tilde{T}_{Ri} + \left\{ \alpha_i W_{Di} (\bar{C}_{PBi} - \bar{C}_{PDi}) + [W_{Di} - (\alpha_i + 1) W_R] \bar{C}_{PBi} \right\} T_{Di}}{\alpha_i W_R \bar{C}_{PRi} \beta_i + \left\{ \alpha_i W_{Di} (\bar{C}_{PBi} - \bar{C}_{PDi}) + [W_{Di} - (\alpha_i + 1) W_R] \bar{C}_{PBi} \right\}}$$

The above equation does not take into account conversion of pumping work into sensible heat.

$$(11) \quad T_{Ri} = \tilde{T}_{Ri} + \frac{\beta_i \left\{ \alpha_i W_{Di} (\bar{C}_{PBi} - \bar{C}_{PDi}) + [W_{Di} - (\alpha_i + 1) W_R] \bar{C}_{PBi} \right\} (T_{Di} - \tilde{T}_{Ri})}{\alpha_i W_R \bar{C}_{PRi} \beta_i + \alpha_i W_{Di} (\bar{C}_{PBi} - \bar{C}_{PDi}) + [W_{Di} - (\alpha_i + 1) W_R] \bar{C}_{PBi}}$$

$$(12) \quad \tilde{W}_{Di} = W_{Di} + \frac{(W_R - W_{Di}) \alpha_i W_R \bar{C}_{PRi} \beta_i (T_{Di} - \tilde{T}_{Ri})}{\alpha_i \bar{\Delta}_i \left\{ \alpha_i W_R \bar{C}_{PRi} \beta_i + \alpha_i W_{Di} (\bar{C}_{PBi} - \bar{C}_{PDi}) + [W_{Di} - (\alpha_i + 1) W_R] \bar{C}_{PBi} \right\}}$$

Note that by defining

$$(13) \quad A_{Ri} = \alpha_i W_R \bar{C}_{PRi} \beta_i$$

$$(14) \quad B_{Ri} = \alpha_i W_{Di} (\bar{C}_{PBi} - \bar{C}_{PDi}) + [W_{Di} - (\alpha_i + 1) W_R] \bar{C}_{PBi}$$

eqs. (10) to (12) become respectively:

$$(15) \quad \tilde{T}_{Di} = \frac{A_{Ri} \tilde{T}_{Ri} + B_{Ri} T_{Di}}{A_{Ri} + B_{Ri}}$$

$$(16) \quad T_{Ri} = \tilde{T}_{Ri} + \frac{\beta_i B_{Ri} (T_{Di} - \tilde{T}_{Ri})}{A_{Ri} + B_{Ri}}$$

EXHIBIT 5 - (Page 5 of 11)

$$(17) \quad \tilde{W}_{Di} = W_{Di} + \left[\frac{(W_R - W_{Di})}{\alpha_i \bar{\Delta}_i} \right] \frac{A_{Ri} (T_{Di} - \tilde{T}_{Ri})}{A_{Ri} + B_{Ri}}$$

Eqs. (15), (16) and (17) are the basic equations for stagewise calculations. However, they cannot be evaluated in a straightforward manner since A_R and B_R both contain average values of physical properties. Therefore, a converging iterative procedure is required to calculate eqs. (15) to (17) simultaneously.

Let the stages of the heat recovery section be numbered 1, 2, ..., N, starting at the brine heater. The heat rejection stages are numbered N+1, N+2, ... For the heat rejection section the following assumptions are made:

1. Cooling water and make-up water may flow through separate tube bundles with different temperatures.
2. The distillate leaving stage N may be withdrawn from the system.

Let

$T_{M,N+i}$ = temp. of make-up water entering stage N+i (°F)

$T_{C,N+i}$ = temp. of cooling water entering stage N+i (°F)

W_M = flow rate of make-up water (lb/hr)

W_C = flow rate of cooling water (lb/hr)

The meaning of all other variables appearing in the following relations is consistent with previous discussion or should be clear from the subscripts used.

EXHIBIT 5 - (Page 6 of 11)

In addition we define:

$$(18) \quad \beta_{M,N+i} = \left\{ 1 - \exp \left[\frac{\bar{U}_{M,N+i} \pi_{M,N+i}^{D} L_{N+i} n_{M,N+i}}{W_M \bar{C}_{PM,N+i}} \right] \right\}$$

$$(19) \quad \beta_{C,N+i} = \left\{ 1 - \exp \left[\frac{\bar{U}_{C,N+i} \pi_{C,N+i}^{D} L_{N+i} \tilde{n}_{C,N+i}}{W_C \bar{C}_{pC,N+i}} \right] \right\}$$

where

$$(20) \quad \bar{U}_{M,N+i} = \frac{U_{M,N+i} (\tilde{T}_{D,N+i} - T_{M,N+i}) + \tilde{U}_{M,N+i} (\tilde{T}_{D,N+i} - \tilde{T}_{M,N+i})}{(\tilde{T}_{D,N+i} - T_{M,N+i}) + (\tilde{T}_{D,N+i} - \tilde{T}_{M,N+i})}$$

and $\bar{U}_{C,N+i}$ has the same form as (20), except for replacing M by C everywhere at the right hand side.

If the make-up and cooling streams have identical temperatures and flow conditions then eqs. (18)-(20) are not necessary and eqs. (15), (16) and (17) are used with the following changes:

$$(21) \quad W_R \longrightarrow W_{RN} = W_M + W_C$$

$$(22) \quad T_{Ri} \longrightarrow T_{R,N+i} = T_{M,N+i} = T_{C,N+i}$$

For the general case and for no withdrawal of distillate at stage N:

$$(23) \quad \tilde{T}_{D,N+i} = \frac{A_M \tilde{T}_{M,N+i} + A_C \tilde{T}_{C,N+i} + B_R T_{D,N+i}}{A_M + A_C + B_R}$$

EXHIBIT 5 - (Page 7 of 11)

where $B_{RN} = \alpha_i W_{Di} (\bar{C}_{PBi} - \bar{C}_{PDi}) + \left[W_{Di} - (\alpha_i + 1) W_R \right] \bar{C}_{PBi}$

$$(24) \quad A_M = \alpha_i W_M \bar{C}_{PM,N+i} \beta_{M,N+i}$$

$$(25) \quad A_c = \alpha_i W_c \bar{C}_{Pc,N+i} \beta_{c,N+i}$$

If make-up and cooling water streams are identical:

$$(26) \quad A_M + A_c = A_{RN}$$

$$(27) \quad \text{and} \quad \tilde{T}_{M,N+i} = \tilde{T}_{c,N+i} = \tilde{T}_{R,N+i}$$

Similarly,

$$(28) \quad \tilde{T}_{M,N+i} = \tilde{T}_{M,N+i} + \frac{\beta_{M,N+i} A_c (\tilde{T}_{c,N+i} - \tilde{T}_{M,N+i}) + \beta_{M,N+i} B_{RN} (\tilde{T}_{D,N+i} - \tilde{T}_{M,N+i})}{A_M + A_c + B_{RN}}$$

$$(29) \quad \tilde{T}_{c,N+i} = \tilde{T}_{c,N+i} + \frac{\beta_{c,N+i} A_M (\tilde{T}_{M,N+i} - \tilde{T}_{c,N+i}) + \beta_{c,N+i} B_{RN} (\tilde{T}_{D,N+i} - \tilde{T}_{c,N+i})}{A_M + A_c + B_{RN}}$$

$$(30) \quad \tilde{W}_{D,N+i} = W_{D,N+i} + \left[\frac{(W_R - W_{D,N+i})}{\alpha_{N+i} \bar{\Delta}_{N+i}} \right] \left[\frac{(A_M + A_c) \tilde{T}_{Di} - A_M \tilde{T}_{M,N+i} - A_c \tilde{T}_{c,N+i}}{A_M + A_c + B_{RN}} \right]$$

Here eqs. (23), (28), (29) and (30) are the basic equations, corresponding to (15), (16), and (17) above.

If the distillate is withdrawn after leaving stage N, the following relations are valid in the heat rejection section:

$$(31) \quad W_{D,N+i} = 0$$

$$(32) \quad W_{B,N+i} = W_{B,N+i} - W_{D,N+i}$$

Let

$$(33) \quad B_{BN} + \alpha_{N+i} W_{D,N+i} (\bar{C}_{PB,N+i} - \bar{C}_{PD,N+i}) + \left[W_{D,N+i} - (\alpha_{N+i} + 1) W_{B,N+i} \right] \bar{C}_{PB,N+i}$$

Eqs. (23), (28), (29) and (30) are used then with B_{RN} replaced by B_{BN} .

Mordechai Avriel

Heat Transfer Calculations

Heat transfer coefficients will be calculated at conditions corresponding to both ends of a stage and an average will be taken.

Assumptions:

1. Constant vapor temperature, \bar{T}_{Di} across stage i.
2. The condensing film temperature has a greater effect upon the tube wall temperature than the bulk recycle temperature.
3. Temp. drop across tube wall negligible for purposes of wall temp. calculations.

The following wall and film temperatures will be used:

1. Wall temperature

$$T_{Wi} = 0.7 \bar{T}_{Di} + 0.3 T_{Ri}$$

and

$$\bar{T}_{Wi} = 0.7 \bar{T}_{Di} + 0.3 \bar{T}_{Ri}$$

Condensing film temperature:

$$T_{Dfi} = \frac{3}{4} T_{Wi} + \frac{1}{4} \bar{T}_{Di}$$

$$T_{Dfi} = \frac{3}{8} \bar{T}_{Ri} + \frac{5}{8} \bar{T}_{Di}$$

EXHIBIT 5 - (Page 10 of 11)

Similarly

$$\tilde{T}_{Dfi} = \frac{3}{8} T_{Ri} + \frac{5}{8} \tilde{T}_{Di}$$

The following film coefficient correlations will be used: ("Transport. Phenomena" Bird, Stewart, Lightfoot, Wiley p. 399)

1. Inside the Tube

$$h_{Ri} = 0.026 \frac{k_{Ri}}{(ID)_i} \left(\frac{(ID)_i U_{Ri} \rho_{Ri}}{\mu_{Ri}} \right)^{0.8} \left(\frac{C_{PRi} \mu_{Ri}}{k_{Ri}} \right)^{1/3} \left(\frac{\mu_{Ri}}{\mu_{Wi}} \right)^{0.14}$$

where k_{Ri} = thermal conductivity of recycle brine, Btu/(hr)(sq. ft)(°F)/ft.

μ_{Ri} = viscosity of recycle brine, lb/(hr)(ft)

μ_{Wi} = viscosity of recycle brine evaluated at the wall temp., lb/(hr)(ft)

ρ_{Ri} = density of recycle brine, lb/ft³

U_{Ri} = velocity of recycle brine, ft/hr

$(I.D.)_i$ = inside diameter at stage i, ft

All physical properties are evaluated at T_{Ri} or T_{Wi}

Similarly,

$$\tilde{h}_{Ri} = 0.026 \frac{\tilde{k}_{Ri}}{(I.D.)_i} \left(\frac{(I.D.)_i \tilde{\mu}_{Ri} \tilde{\rho}_{Ri}}{\tilde{\mu}_{Ri}} \right)^{0.8} \left(\frac{\tilde{C}_{PRi} \tilde{\mu}_{Ri}}{\tilde{k}_{Ri}} \right)^{1/3} \left(\frac{\tilde{\mu}_{Ri}}{\tilde{\mu}_{Wi}} \right)^{0.14}$$

where the variables have the same meanings as before, but calculated at temps. corresponding to \tilde{T}_{Ri} or T_{Wi} .

EXHIBIT 5 - (Page 11 of 11)

2. Condensing film coefficient

The following correlation will be used: ("Unit Operation", Brown, p. 449)

$$h_{Di} = 0.725 \left[\frac{\left(\frac{k_{Dfi}}{N_i D_i} \right)^3 \left(\frac{\rho_{Dfi}}{\mu_{Dfi}} \right)^2 g \tilde{\lambda}_i}{(T_{Di} - T_{Wi})} \right]^{1/4}$$

where

$\tilde{\lambda}_i$ = latent heat of vaporization at T_{Di}

All other physical properties are calculated at temp. T_{Dfi}

N_i = no. of tubes in a vertical column of tubes.

Similarly,

$$\tilde{h}_{Di} = 0.725 \left[\frac{\left(\frac{\tilde{k}_{Dfi}}{N_i D_i} \right)^3 \left(\frac{\rho_{Dfi}}{\tilde{\mu}_{Dfi}} \right)^2 g \tilde{\lambda}_i}{(T_{Di} - \tilde{T}_{Wi})} \right]^{1/4}$$

Overall heat transfer coefficient, based on outside diameter:

$$U_i = \frac{1}{\frac{1}{h_{Di}} + \frac{D_i}{h_{Ri} (ID)_i} + \frac{\left[\frac{D_i - (ID)_i}{D_i + (ID)_i} \right] \frac{D_i}{(k_m)}} + FF}$$

where k_m = thermal conductivity of tube metal, Btu/(hr)(sq. ft)(°F)/ft

FF = fouling factor, (hr)(sq.ft)(°F)/BTU

Similarly,

$$\tilde{U}_i = \frac{1}{\frac{1}{\tilde{h}_{Di}} + \frac{D_i}{\tilde{h}_{Ri} (ID)_i} + \frac{\left[\frac{D_i - (ID)_i}{D_i + (ID)_i} \right] \frac{D_i}{(k_m)}} + FF}$$

FF was linear with temperature, °F, varying from .0005 at 85°F to

0.001 $\frac{ft^2, hr, °F}{Btu}$ at 250°F.

Exhibit 6.

Operating Experience at San Diego Flash Distillation Plant. Reprinted from paper by Foster and Herlily of Burns and Roe presented at First International Symposium on Water Desalination October 3-9, 1965

T A B L E 1
OPERATING DATA ANALYSIS

Stage No.	Temperature °F of Flash Dist Brine	Temp Term	Difference Log Mean	Heat Trans MBTU/Hr	Overall U From Plant Data	Theor. U Clean for 1 Tube	** See Note Below	Flash Brine MB/Hr	Distill. Product MB/Hr	Recycle Brine MB/Hr	Concent. Wt. Pent	Product. Lb/Hr	Press. Psia	RFE °F	Vapor Flow M Cu Ft/Sec.
1	196.3	194.0	190.3	3.70	5.206	5648.56	692.4	1298.7	0.006741	1779.2	5.75	1785.0	6.320	5754.3	61.22
2	192.7	190.2	186.9	3.30	4.859	5949.60	781.4	1280.2	0.004985	1773.2	11.77	1785.0	6.341	6024.4	69.04
3	189.0	186.4	183.3	3.05	4.663	6217.27	850.8	1262.0	0.003828	1766.9	18.03	1785.0	6.364	6258.1	77.36
4	185.2	182.6	179.6	2.95	4.596	6396.49	888.2	1244.9	0.003226	1760.5	24.43	1785.0	6.387	6400.2	85.46
5	181.3	178.8	175.8	2.94	4.613	6499.73	899.2	1228.9	0.002982	1754.0	30.90	1785.0	6.410	6464.2	93.37
6	177.4	174.9	172.0	2.99	4.674	6539.55	892.9	1213.7	0.002959	1747.6	37.36	1785.0	6.434	6466.0	101.13
7	173.5	171.1	168.1	3.06	4.747	6528.47	877.7	1199.1	0.003052	1741.2	43.77	1785.0	6.458	6412.7	108.80
8	169.6	167.3	164.2	3.13	4.807	6478.07	860.0	1185.0	0.003188	1734.8	50.10	1785.0	6.481	6322.8	116.47
9	165.8	163.5	160.3	3.17	4.838	6398.88	844.1	1171.4	0.003309	1728.6	56.30	1785.0	6.505	6205.2	124.23
10	162.0	159.7	156.5	3.18	4.829	6300.07	832.5	1158.2	0.003377	1722.6	62.37	1785.0	6.528	6069.6	132.19
11	158.3	155.9	152.7	3.16	4.779	6189.22	826.6	1145.2	0.003366	1716.7	68.29	1785.0	6.550	5923.8	140.46
12	154.6	152.1	149.0	3.10	4.687	6072.28	826.7	1132.5	0.003265	1710.9	74.07	1785.0	6.572	5774.2	159.15
13	151.1	148.4	145.4	3.01	4.562	5953.47	832.7	1119.9	0.003078	1705.3	79.69	1785.0	6.594	5625.1	168.35
14	147.6	144.7	141.8	2.89	4.413	5835.38	843.9	1107.5	0.002819	1699.8	85.17	1785.0	6.615	5479.2	178.12
15	144.2	141.1	138.3	2.76	4.250	5718.91	858.7	1095.2	0.002513	1694.4	90.51	1785.0	6.636	5337.8	188.50
16	140.8	137.5	134.9	2.63	4.086	5603.71	875.1	1083.1	0.002194	1689.2	95.71	1785.0	6.656	5200.5	189.45
17	137.5	134.1	131.6	2.51	3.935	5488.12	890.1	1071.2	0.001899	1684.2	100.78	1785.0	6.676	5066.0	200.89
18	134.3	130.7	128.3	2.42	3.806	5369.54	900.4	1059.6	0.001668	1679.2	105.71	1785.0	6.696	4931.7	212.65
19	131.2	127.4	125.1	2.35	3.708	5244.93	902.5	1048.3	0.001540	1674.4	110.50	1785.0	6.715	4794.8	224.49
20	128.1	124.2	121.9	2.33	3.649	5111.00	933.8	1037.5	0.001549	1669.8	115.16	1785.0	6.734	4652.2	236.07
21	125.2	121.2	118.9	2.34	3.630	4964.80	972.9	1027.2	0.001720	1665.3	119.66	1785.0	6.752	4500.8	247.03
22	122.3	118.3	115.9	2.39	3.649	4804.19	940.3	1017.4	0.002072	1660.9	124.00	1785.0	6.770	4338.5	256.94
23	119.5	115.5	113.0	2.48	3.700	4628.52	798.3	1008.4	0.002609	1656.8	128.16	1785.0	6.787	4164.5	265.45
24	116.8	112.8	110.2	2.60	3.775	4439.01	750.5	1000.0	0.003324	1652.8	132.14	1785.0	6.803	3979.4	272.27
25	114.3	110.3	107.6	2.72	3.860	4239.89	700.9	992.3	0.004189	1649.0	135.93	1785.0	6.819	3786.9	277.38
26	111.9	107.9	105.0	2.85	3.944	4038.68	653.4	985.1	0.005151	1645.4	139.52	1785.0	6.834	3593.5	281.06
27	109.6	105.6	102.6	2.96	4.012	3847.37	611.9	978.3	0.006118	1642.0	142.93	1785.0	6.848	3409.6	284.14
28	107.3	103.4	100.3	3.04	4.052	3683.04	580.0	971.5	0.006947	1638.8	146.18	1785.0	6.861	3250.4	288.13
29	105.2	101.2	98.1	3.08	4.058	3568.97	561.3	964.0	0.007442	1635.6	149.32	1785.0	6.875	3136.9	295.53
30	103.1	99.0	96.0	3.06	4.026	3535.55	560.4	955.2	0.007375	1632.5	152.41	1785.0	6.888	3096.3	310.09
31	100.9	96.8	93.9	2.98	3.966	3621.38	582.6	944.1	0.006570	1629.4	155.58	1785.0	6.901	3163.4	337.32
32	98.6	94.5	91.7	2.85	3.900	3874.09	633.9	929.6	0.005016	1626.0	158.96	1785.0	6.915	3381.1	385.21
33	96.0	92.1	89.4	2.70	3.864	4351.69	718.7	911.0	0.002935	1622.2	162.76	1785.0	6.932	3801.3	485.42
34	92.8	89.3	86.8	2.58	3.918	5123.24	834.5	887.8	0.000720	1617.7	167.25	1785.0	6.951	4485.5	595.29
Heat Rejection Stages															
Seawater															
35	89.0	86.2	82.5	3.79	5.874	3968.91	384.4	556.3	0.008038	1612.2	169.76	1365.3	6.975	2512.5	365.83
36	84.1	82.7	77.6	5.02	12.388	16295.70	748.4	541.5	0.005102	1606.0	175.42	1365.3	7.001	5666.0	920.11
Total															
Production = 358,800, Lb/Hr															
Temp of Flashing Brine from Brine Heater = 57.9 °F															
Seawater Temp = 57.9 °F															
Heat Recovery Stages A-Side															
Recycle Ratio A-Side 4.97															
Concent. Ratio A-Side 1.80															
Heat Rejection Stages															
Seawater															
A-Side															
B-Side															
Total															
Number of Tubes															
Tube Material															
Tube Length, Ft															
Velocity of Recycle Brine, Ft/Sec															
Temperature of Recycle entering, °F															
Temperature Rise of Recycle Brine, °F															
Log Mean Temperature Difference															
Overall Heat Transfer Coefficient, Btu/Ft ² /Hr															
Theoretical Overall U for Clean Tubes															
Heat Requirement, MBTU/Hr															
Steam Requirement, Lb/Hr															
Production Rate (Recover and Reject Stages), Lb/Hr															
Steam Economy Ratio Lb Product/Lb Steam															

** Difference between Reciprocal of Overall U from Plant Data and Reciprocal of U Clean

ECL 70

SMD/46

BECHTEL CORPORATION (C)

Process Analysis and Optimization of a Desalting Plant Design

After studying the multistage flash process Dr. Wilde suggested that the "Pontryagin Maximum Principle" might be a useful optimization tool. This method is a sophisticated application of variational calculus and had been examined by Dr. Wilde at Stanford. Crawford and Avriel found it necessary, as in most optimization processes, to develop explicit relationships for the process variables from the cycle analysis so that partial derivatives could be conveniently determined. Crawford summarized (Exhibit 7) his energy and material balance equations for the computer program.

Initially Rufus decided to base his computations on cylindrical outer shell evaporators and worked with this geometry for about one month. Since previous experience at San Diego indicated that a rectangular shape allowed a larger volume of brine flow, Clark and Galstaun suggested that Crawford switch geometries. The original rationale for choosing a cylindrical cross-section was based on the loading of the shell when under vacuum. However, the magnitude of the loads actually encountered were such that no serious stress problems developed in the San Diego rectangular shells. Some question existed as to whether the cylindrical or rectangular shape was the more economical evaporator configuration.

The cost estimates for this program were based on steel pressure vessels. Earlier, Bechtel had made a study comparing coated concrete and steel shells. Although the cost estimates indicated a competitiveness for concrete, limited experiments indicated that the surface coatings would not stand up under exposure to hot brine water. It was assumed in the study that if concrete were used for the pressure vessels, the slabs would be cast at the construction site. Crawford said, "At the present time the absence of data on the concrete application makes it difficult to predict the long-term reliability of these evaporator vessels."

The tube material and its thermodynamic properties were inputs to the computer program. Aluminum-brass was used since this provided a check on the 4427 study. However, based on power plant experience the general feeling is that if they are going to design for a 30 year life that 70-30 Cu-Ni will have to be used. Crawford said, "Cu-Ni is probably the best material you can afford to use." The tube diameter is an input to the computer program and 5/8, 3/4, 7/8 and 1 inch are standard, and currently the maximum mill tube length is 60 feet. The 3/4 inch diameter (O.D.) is most common and is usually assumed for the first try.

The question of deareation arose repeatedly but the decision was made to include it only as a fixed cost item. Crawford though to himself, "This would be an optimization study in itself." In estimating the cost of deareation it was included as an integral portion of the heat rejection section.

For scale control Crawford was told by Galstaun to design so that either polyphosphate or sulfuric acid could be used and these would set the temperature limitations of 195° and 250°F, respectively. Crawford says that if anyone comes up with a new method of scale control that would remove these temperature limitations, new cost data could be fed into the computer program. As for the inherent drop in performance with time, this was taken care of as a fouling factor. The fouling factor is assumed to be a function of the temperature and the condenser tube material as described in part (B).

On the question of drives, Crawford said he made the decision to limit the study to electric drives since cost data were more readily available than for other systems and this seemed to have the approval of Galstaun. Steam drives were not considered even though they were known to have been used on the main pumps of the San Diego plant. Crawford decided to wait until data on the steam drives became available before incorporating it into their optimization model.

Crawford discussed the philosophy of cost estimation. He noted that the "broad brush" treatment often gave answers within 5 to 10% of detailed cost analysis and in much less time. Therefore since the refined equations are proprietary, Rufus offered some of the broad brush guide lines for establishing costs.

Crawford noted that steam costs average from 20 to 30 cents per million Btu at 250°F and would be more expensive at higher temperatures. However, to lower the steam temperature at the brine heater more surface area is required.

In sizing heaters and other extensive equipment he said it's possible to roughly estimate material costs on a lineal or square foot basis. "For example, you can figure something under \$1.00/ft² of heat transfer surface to which must be added the cost of the tubes \$2.80/ft² Cu-Ni or \$2.30 Al-Brass," said Crawford. A large brine heater may then run \$4.50/ft² if the foundations, piping and removal of non-condensibles are included. The vessels which contain the tubes might run from \$20 to \$30 per ft² of evaporator bottom surface area including auxiliaries. A typical height of these vessels is 8 feet, 40 to 60 feet long and 50 feet wide.

Crawford, off the top of his head, set down in this fashion many direct and indirect operating costs. A typical breakdown (Exhibit 8) is reproduced from Bechtel's report to the Office of Saline Water (OSW No. 116). In a similar manner for any given set of boundary conditions (and with greater accuracy by computer input), Rufus was able to obtain a final cost for producing water given as cents per thousand gallons.

Assignment

1. Outline how you would attack the problem of process optimization.
2. Crawford stated that there are a number of approaches to cycle analysis,
 - (a) Assume equal temperature drop per stage.
 - (b) Assume equal condenser area per stage.
 - (c) Assume equal distilled water production per stage.

(But that it doesn't make much difference which one you use)

How do you think Crawford came to this conclusion?

How else might you get started?
3. From the data in Exhibit 8 derive cost relations for use in the computer program. If these data are not sufficient, what additional data did Crawford search for?



BECHTEL CORPORATION

INTER-OFFICE MEMORANDUM

To M. G. Martin

Date November 3, 1965

From R. W. Crawford

Dept. Scientific Development

Copies to L. S. Galstaun
R. L. Clark
M. Avriel

At 301 Mission

Subject: Final Relations for Heat and Material Balances for Optimization Project Program.

I have found it necessary to include one additional input for solution of the MSF overall heat and material balances. The memorandum that was sent to you dated November 2, 1965 should be destroyed. The following procedure should be programmed.

1. Inputs to program:
 - a. Plant Performance Ratio.
 - b. Brine Temperature at Heat Outlet (T_{B1})
 - c. Product Flow Rate and Temperature (W_p , T_p).
 - d. Concentration of Recycle Brine (C_R)
 - e. Temperature Difference Between Exiting Makeup and Exiting Flashing Brine (ΔT_M).
 - f. Temperature Difference between Exiting Coolant and Exiting Flashing Brine (ΔT_C).
 - g. Concentration and Temperature of Sea (C_s , T_s).
2. Using values for $T_{B,1}$, T_p , W_p and C_R , calculate $C_{B,K}$, $T_{B,K}$ and W_R as described in an earlier memorandum on recycle rate calculations.
3. With the plant production (W_p), calculate the total heat duty at the brine heater using the performance ratio given for the plant.

$$Q = W_p / PR$$

where: PR = Performance Ratio (lb. Product/1000 Btu)

4. Calculate the brine temperature entering the brine heater ($T_{R,1}$) as follows:
 - a. With $T_{B,1}$ and C_R , calculate the heat capacity of the brine, $\overline{C_p}$.
 - b. Solve for $T_{R,1}$ (exiting recycle temperature from stage 1) as below.

$$T_{R,1} = T_{B,1} - \frac{Q}{W_R \overline{C_p}}$$

- c. With C_R and $(T_{B,1} + T_{R,1})/2$ (the average brine temperature in the brine heater) calculate a new heat capacity of the brine ($\overline{C_p}$).
- d. Repeat Steps b and c until the difference between values of $T_{R,1}$ on two successive iterations $\leq 0.005^\circ\text{F}$.

5. Calculate the blowdown flow.

$$W_{BL} = \frac{W_p}{((C_R/C_S) - 1)}$$

6. Calculate the makeup flow.

$$W_M = W_p + W_{BL}$$

7. Calculate the enthalpy of the blowdown initial recycle as:

$$h_{BL} = h_{R,N} = \frac{W_{B,K} h_{B,K} + W_M h_{M,J}}{W_{BL} + W_R}$$

8. Substitute $h_{R,N}$ into the heat recovery section heat balance to solve the total energy leaving the heat recovery section with the distillate and brine streams.

$$(W_{D,J} h_{D,J} + W_{B,J} h_{B,J}) = W_R (h_{R,N} + h_{B,1} - h_{R,1})$$

9. With ΔT_C and $T_{B,K}$, calculate $T_{C,J}$ and $h_{C,J}$.
10. Substitute values obtained into heat rejection section heat balance obtain coolant flow rate.

$$W_C = \frac{h_{R,N}(W_{BL} + W_R) - (W_{D,J} h_{D,J} + W_{B,J} h_{B,J}) - W_M h_{M,J} + W_{php}}{h_S - h_{C,J}}$$

11. Start heat recovery section stagewise calculations. Terminate the section when the recycle temperature for a stage i equals $T_{R,N}$.
Test necessary for each stage calculated.
12. Calculate heat rejection section.

It has been suggested that a possible alternative to specifying an outlet coolant temperature is determination of a coolant flow based upon the size of the recycle flow. That is:

$$W_C = XW_R$$

This alternative should be programmed. The input to the program would not require ΔT_C , but would instead require the percentage factor to be used for calculation of the coolant flow. The following changes will be necessary in the basic calculation procedure:

9. (Alternate) With the percentage factor, calculate the coolant flow rate knowing the recycle flow.
10. (Alternate) Substitute values obtained into heat rejection section heat balance to obtain outlet coolant enthalpy.

$$h_{C,J} = \frac{W_M h_S - W_{php} + (W_{D,J} h_{D,J} + W_{B,J} h_{B,j}) - h_{R,N} (W_{BL} + W_R) + h_S}{W_C}$$


R.W. CRAWFORD

Scientific Development Department

RWC:pg

Exhibit 8* (1)

OPERATING COST

Capacity: Power 134 mw, net
water 50x10⁶ GPD
Investment: \$60,910,000
Load Factor: 90.4%

	<u>Units</u>	<u>\$/Unit</u>	<u>Units/D</u>	<u>\$/Op. Day</u>
A. Direct Operating Cost				
1. Fuel	10 ⁶ Btu	0.20	51,240.	10,248
2. Chemicals				
Sulfuric Acid	Tons	20.00	65.7	1,314
Chlorine	lbs	0.0635	250.	16
Others				1
3. Supplies and Maint. Material (0.0015% of Invest./Day)				914
Subtotal				12,493
4. Labor				
Operating, Power	Men	27.96	15.	419
Water	Men	27.96	12.	336
Maintenance (0.0015% of Invest/Day)				914
Total Direct Labor				1,669
Payroll Extras, 15% of Direct Labor				250
Total Labor Payroll				1,919
TOTAL DIRECT OPERATING COST				14,412
B. Indirect Operating Cost				
5. General and Admin. Overhead (30% of Payroll)				576
6. Amortization (0.0224% of Capital)				13,600
7. Taxes and Insurance (0.006% of Capital)				3,650
TOTAL INDIRECT OPERATING COST				17,826
C. Total Direct and Indirect Operatint Cost				32,238
8. Interest on Working Capital (0.725% of "C")				234
TOTAL OPERATING COST				32,472
D. Electric Power Credit, 134 mw @ \$3.11/mwh				10,000
E. Net Cost of Water				22,472
Net Cost of Water \$ per Thousand Gallons				0.449

ESTIMATE SUMMARY

ITEM		Thousands of Dollars		
		PLANT NO. 1		Notes
		Water	Power	
I	Materials	20,838	10,495	(1)
	Subcontracts	3,620	6,190	(2)
	Labor	<u>1,365</u>	<u>1,795</u>	(3)
	Total Direct Cost	25,823	18,480	(4)
II	Contractor's Field Office Services	765	920	(5)
III	Startup Cost	150	150	(6)
IV	Total Field Cost	26,738	19,550	
V	Contractor's Home Office Services, Incl.Engineering, Procurement and Management	3,600	3,000	(7)
VI	Land for Plant and Canal	60	20	(8)
VII	Contingency	<u>3,000</u>	<u>1,500</u>	(9)
VIII	Total Construction Cost	33,398	24,070	
IX	Interest During Construction	<u>2,002</u>	<u>1,440</u>	(10)
X	Total Plant Investment	<u>35,400</u>	<u>25,510</u>	(11)
XI	Total Plant Investment (Both Plants)	60,910		

WATER PLANT
ESTIMATE IN \$M

	<u>Quantity</u>	<u>Materials</u>	<u>Subc'ts.</u>	<u>Labor</u>	<u>Total</u>
Evaporators					
Concrete Structure	36 Stages		\$ 2,520		
Condenser Bundles	7,050,000 SF	\$14,800		\$ 286	
Internal Piping & Baffles	432 Sets	1,540		141	
Insulation, Lining & Paint			415		
Additions for Heat Rejection	2 Stages	<u>100</u>	<u>50</u>	<u>13</u>	
Subtotals		\$16,440	\$ 2,985	\$ 440	\$19,865
Sulfuric Acid Tank	10,000 bbl		20		
CO ₂ Release Tank	50' x 200'		75		
Chlorine Tank and Accessories		20		2	117
Brine Heaters (3)	75,600 SF	302		2	
Product Coolers (3)	34,800 SF	139		1	
Condensers(3)(Ejector)		40		1	485
Recycle Brine Pump	15,000 HP	1,050		10	
Other Pumps (14)	7,186 HP	620		15	1,695
Ejectors	3 Stages	15		1	16
Air Compressors and Accessories (2)	1,200 SCFM	<u>50</u>		<u>4</u>	<u>54</u>
TOTAL MAJOR EQUIPMENT		\$18,656	\$ 3,080	\$ 476	\$22,412
Instruments		\$ 165		\$ 32	\$ 197
Piping & Valves (External		1,000	30	180	1,210
Intake Structure		150		125	
Intake Structure Equipment and Controls		200		40	
Outlet Canal			60		
Ejector Structures		7		2	
Pipe Support Structures		25		25	634
Insulation			135		135
Electricals		400		110	510
Foundations(excl.Evap.Cells)		75		100	
Piling (Under Pump Bays)			10		185
Sewers and Paving		10	15	15	40
Painting (excl. Evap.)			<u>20</u>		<u>20</u>
TOTAL - OTHER MATERIALS		\$ 2,032	\$ 270	\$ 629	\$ 2,931
PLANT TOTALS		\$20,688	\$3,350	\$1,105	\$25,143
Add Prorated Portion of Common Facilities		<u>150</u>	<u>270</u>	<u>260</u>	<u>680</u>
TOTAL DIRECT COST		\$20,838	\$3,620	\$1,365	\$25,823